

## APPENDICES

### APPENDIX-A

Appendix-A represents various data obtained from the experimental work. Figures 1 and 2 are the calibration curves for air and water rotameters. Figure 3 compares Axial holdup with total holdup. Figures 4 through 7 represent holdup variation with height at increasing superficial gas velocities for porous and sieve plates. Figures 8 through 12 are the gas disengagement curves for different CMC concentrations. Figure 13 is a parity plot for CMC solutions for a porous plate distributor in bubble flow. It compares the calculated holdup from the empirical correlation with the observed holdup.

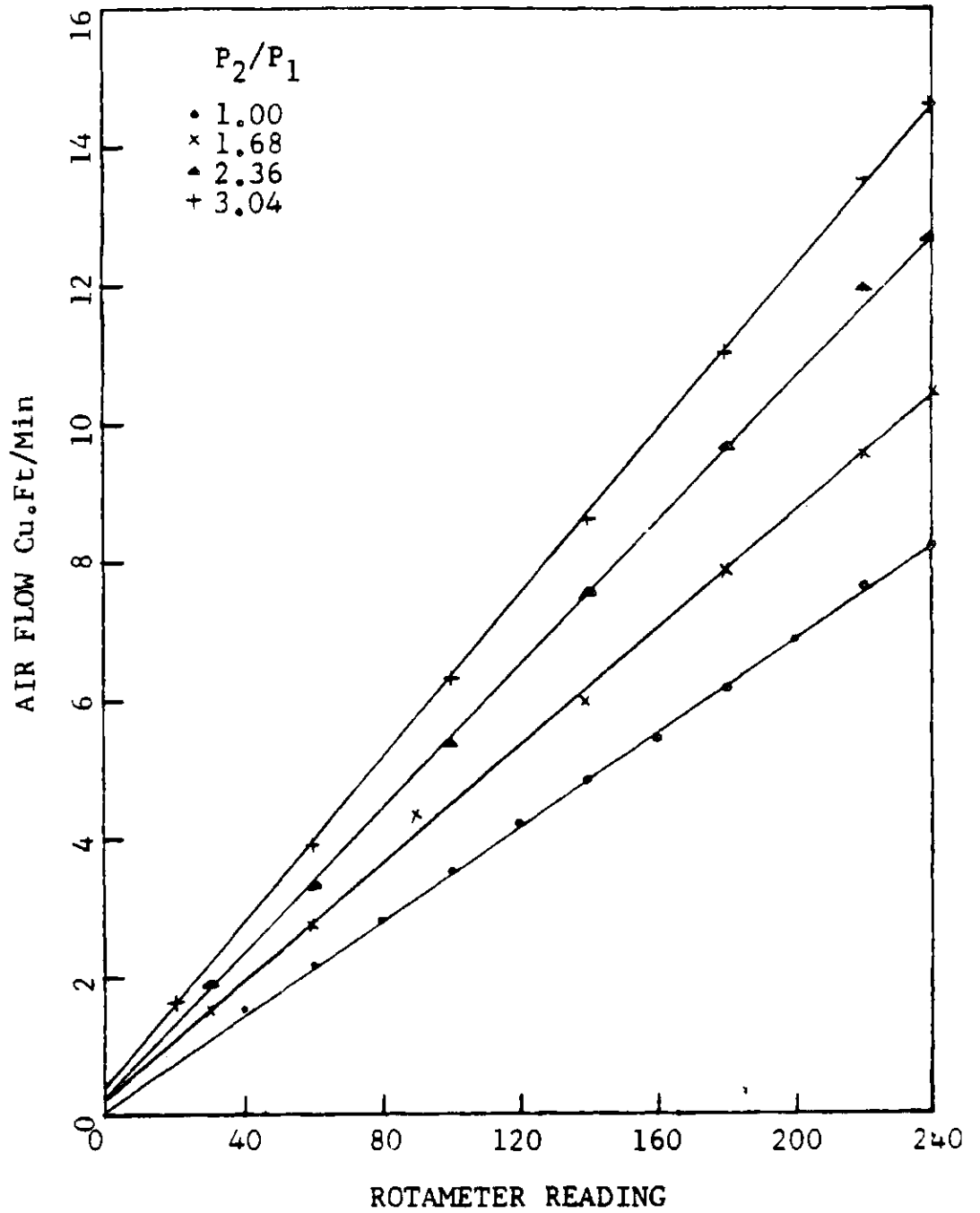


FIG. 1 CALIBRATION FOR AIR FLOW METER  
INLET PRESSURE = 30 Psig

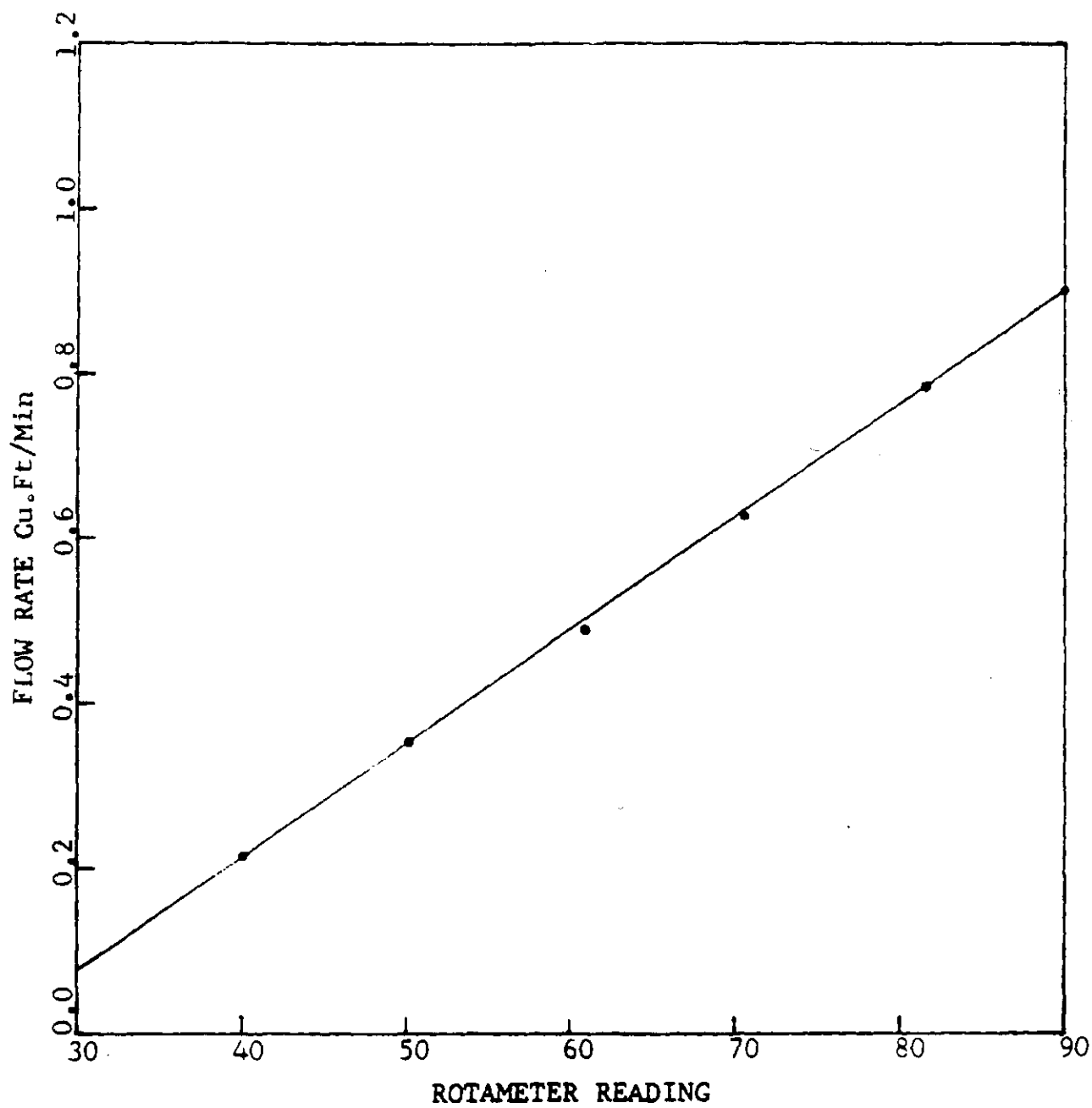


FIG. 2 CALIBRATION FOR WATER FLOW

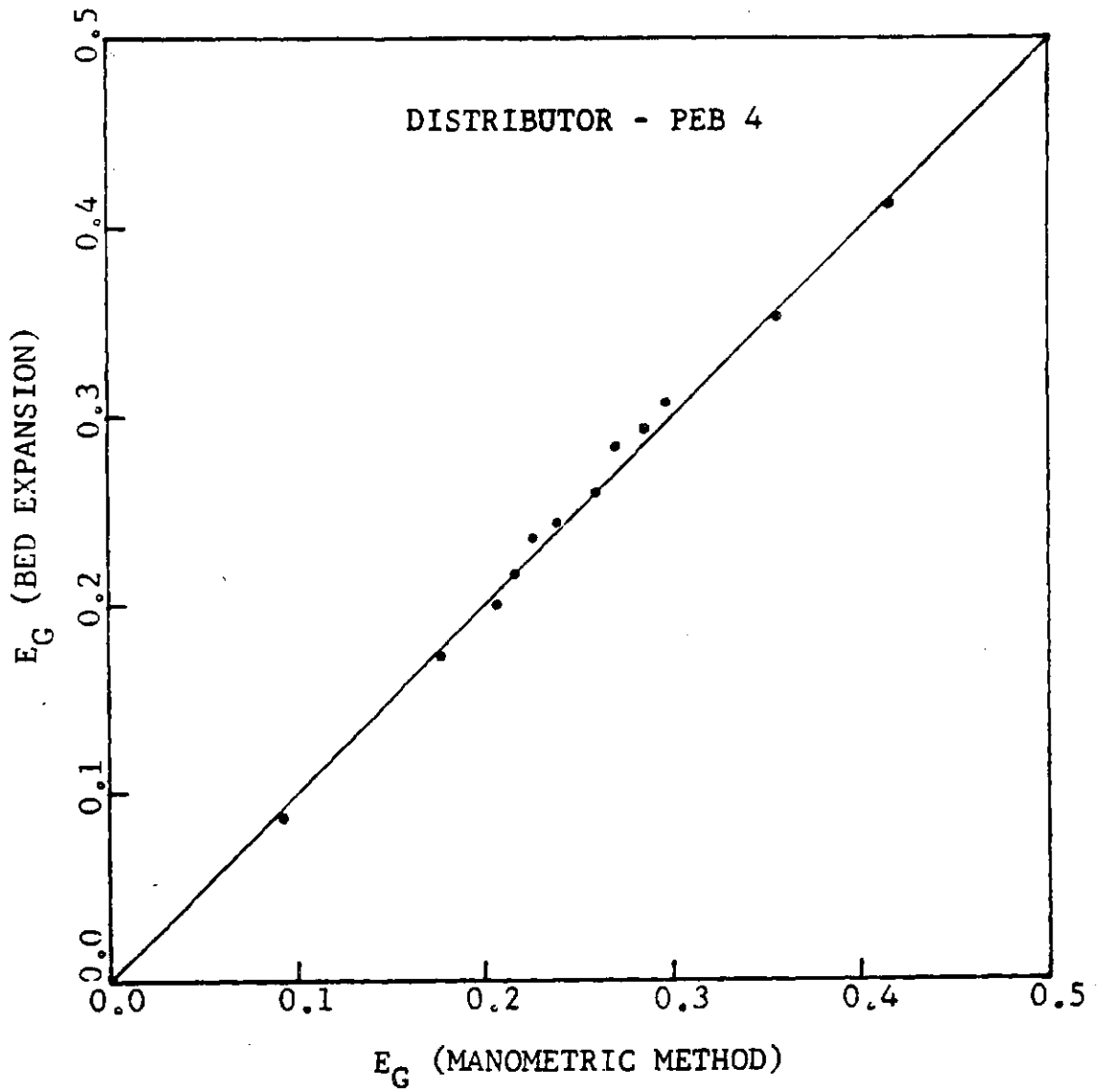


FIG. 3 GAS HOLDUP-BED EXPANSION VS MANOMETRIC METHOD

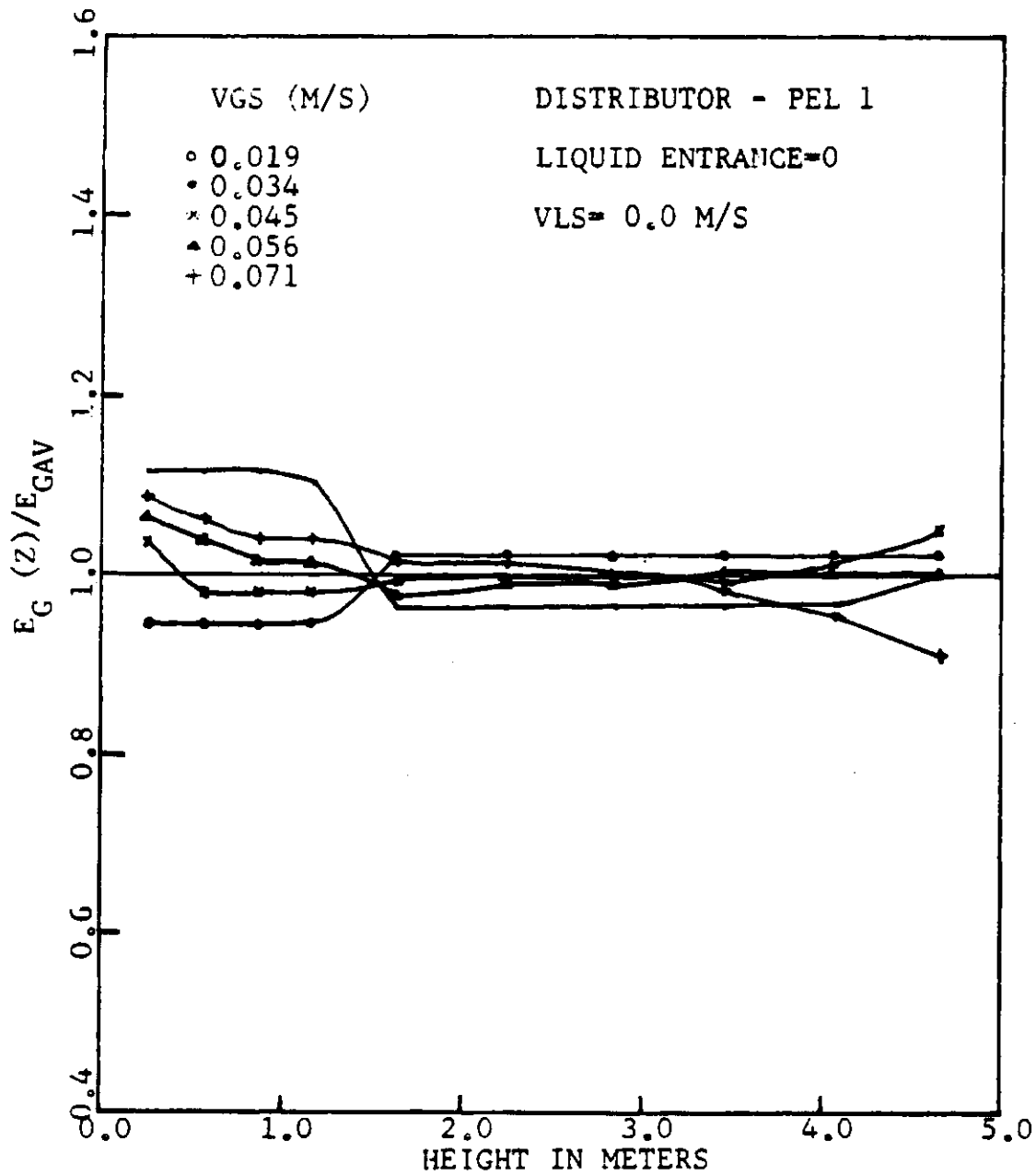


FIG. 4 HOLDUP VARIATION WITH HEIGHT - BUBBLE FLOW

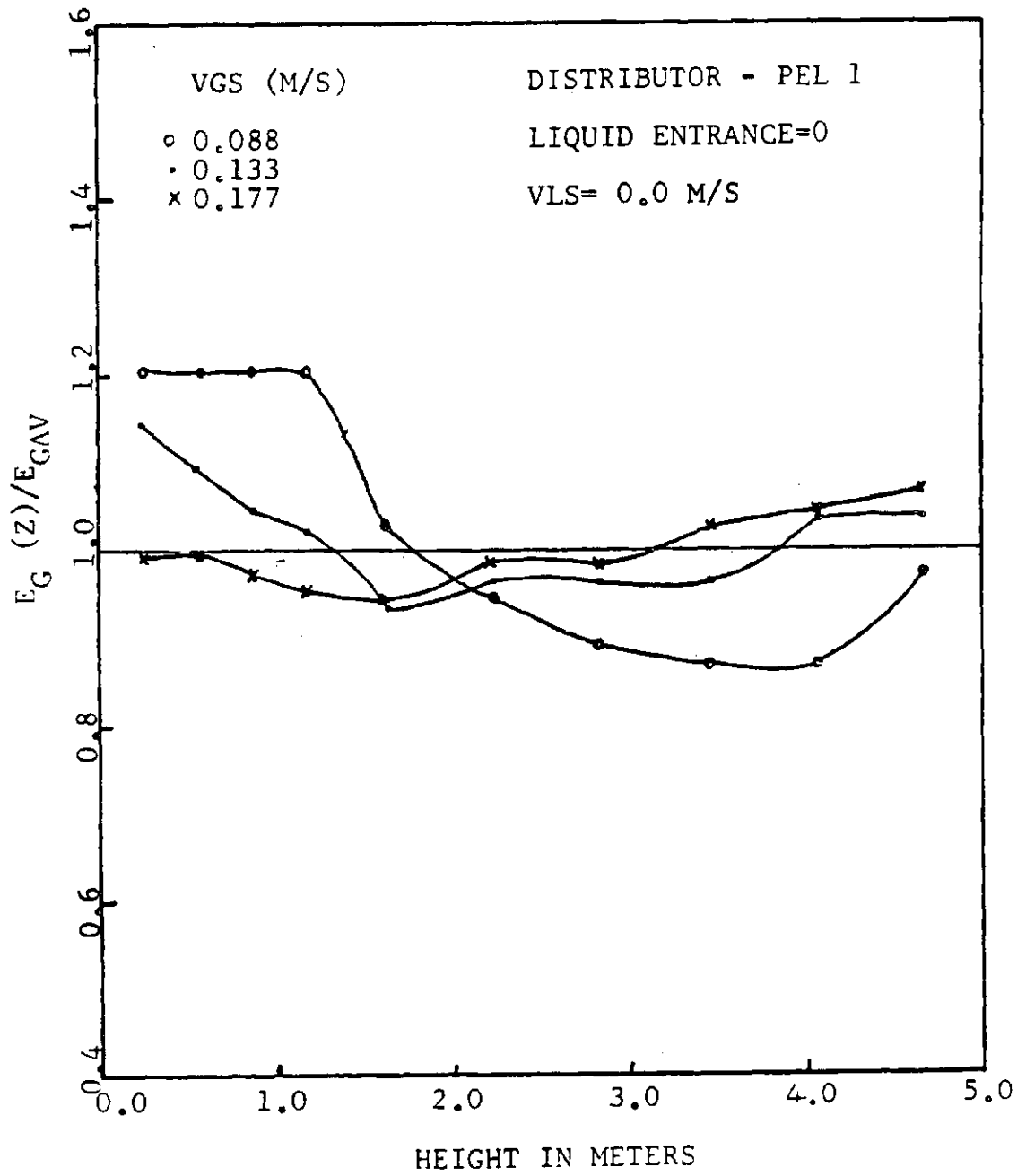


FIG. 5 HOLDUP VARIATION WITH HEIGHT - BUBBLE - SLUG FLOW

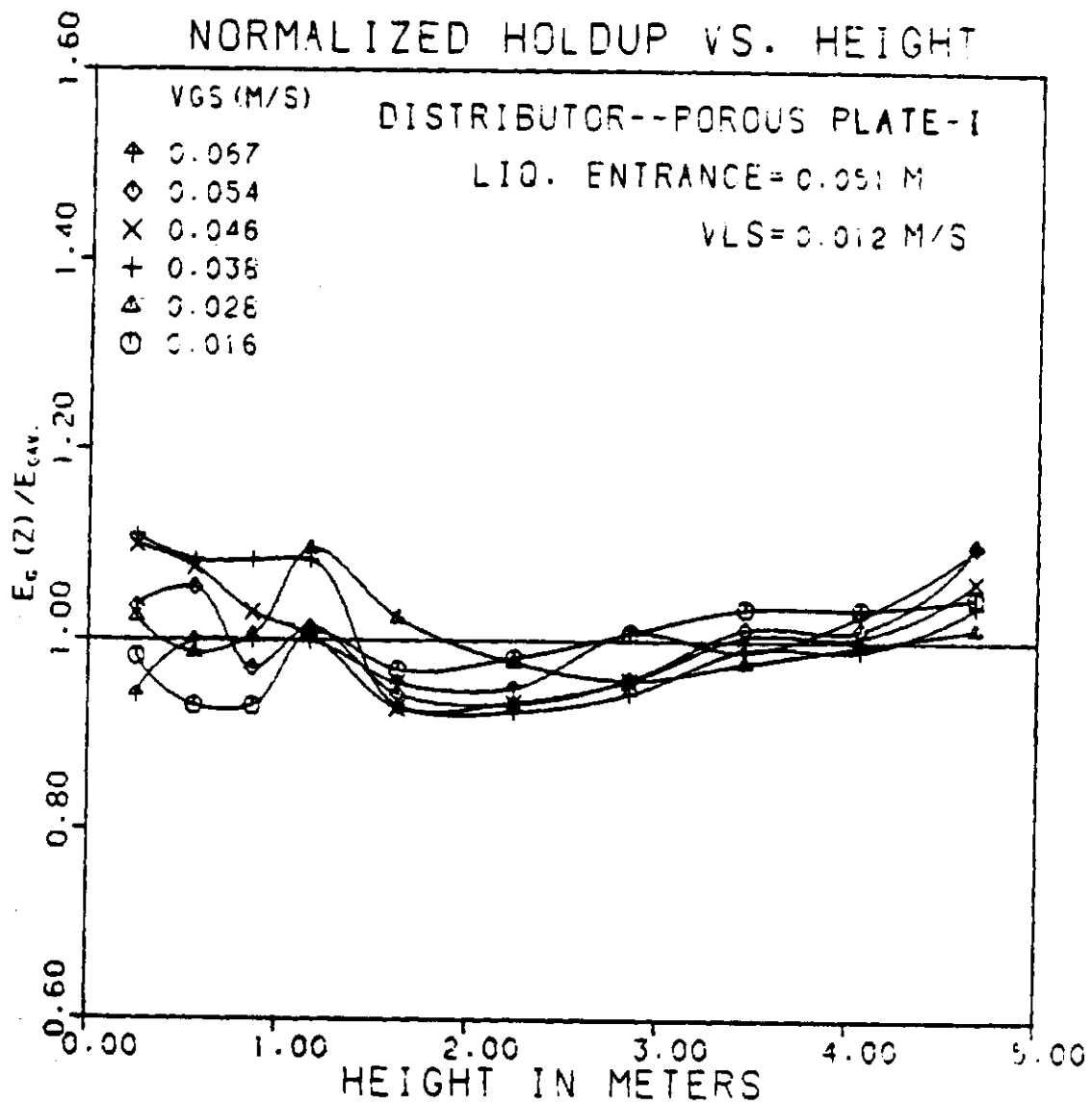


FIG. 6 HOLDUP VARIATION WITH HEIGHT AT LOW GAS VELOCITIES  
POROUS PLATE-I

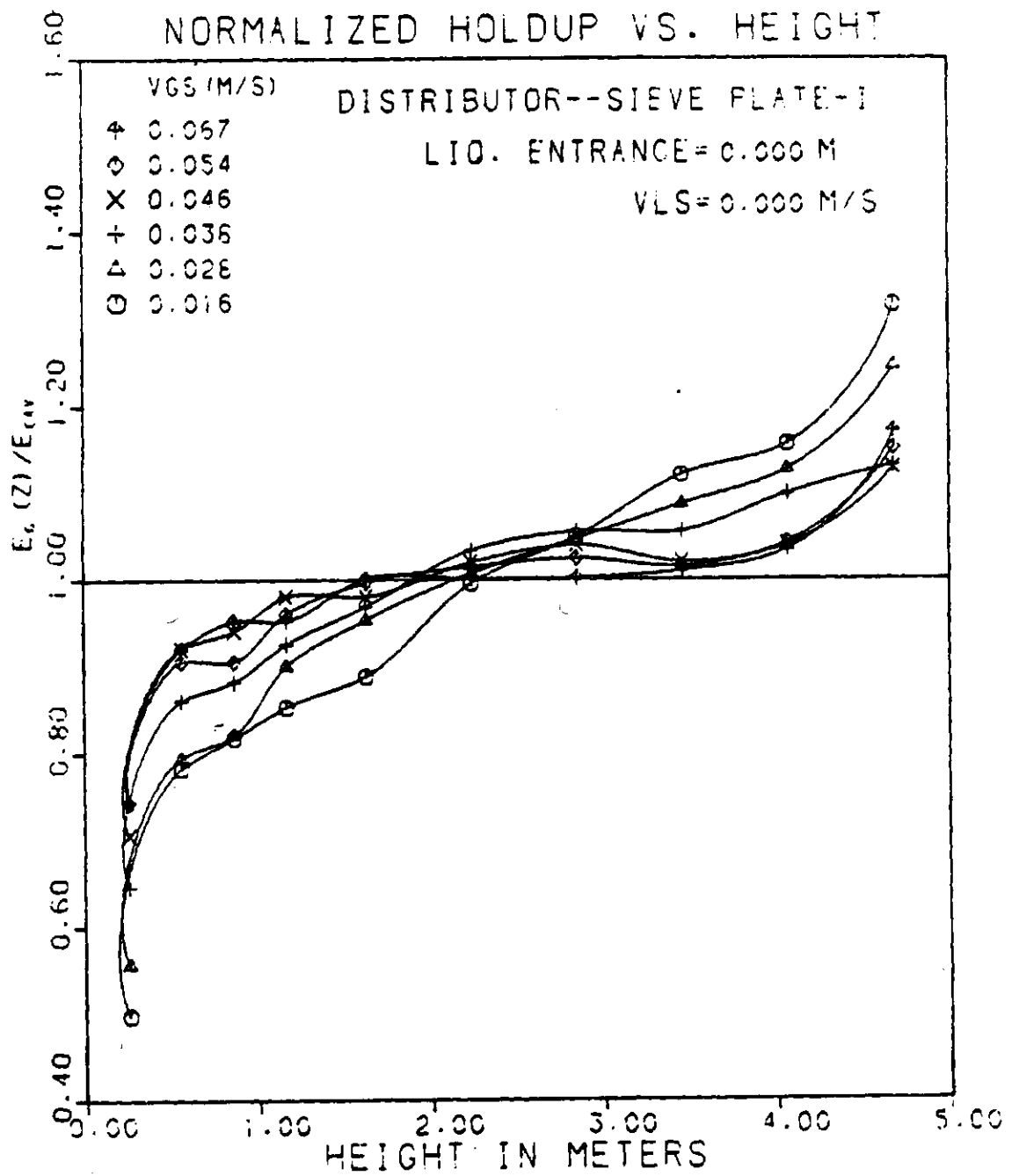


FIG. 7 HOLDUP VARIATION WITH HEIGHT AT LOW GAS VELOCITIES  
SIEVE PLATE-I



FIGURE 8

GAS DISENGAGEMENT FOR  
A 0.25% CMC SOLUTION

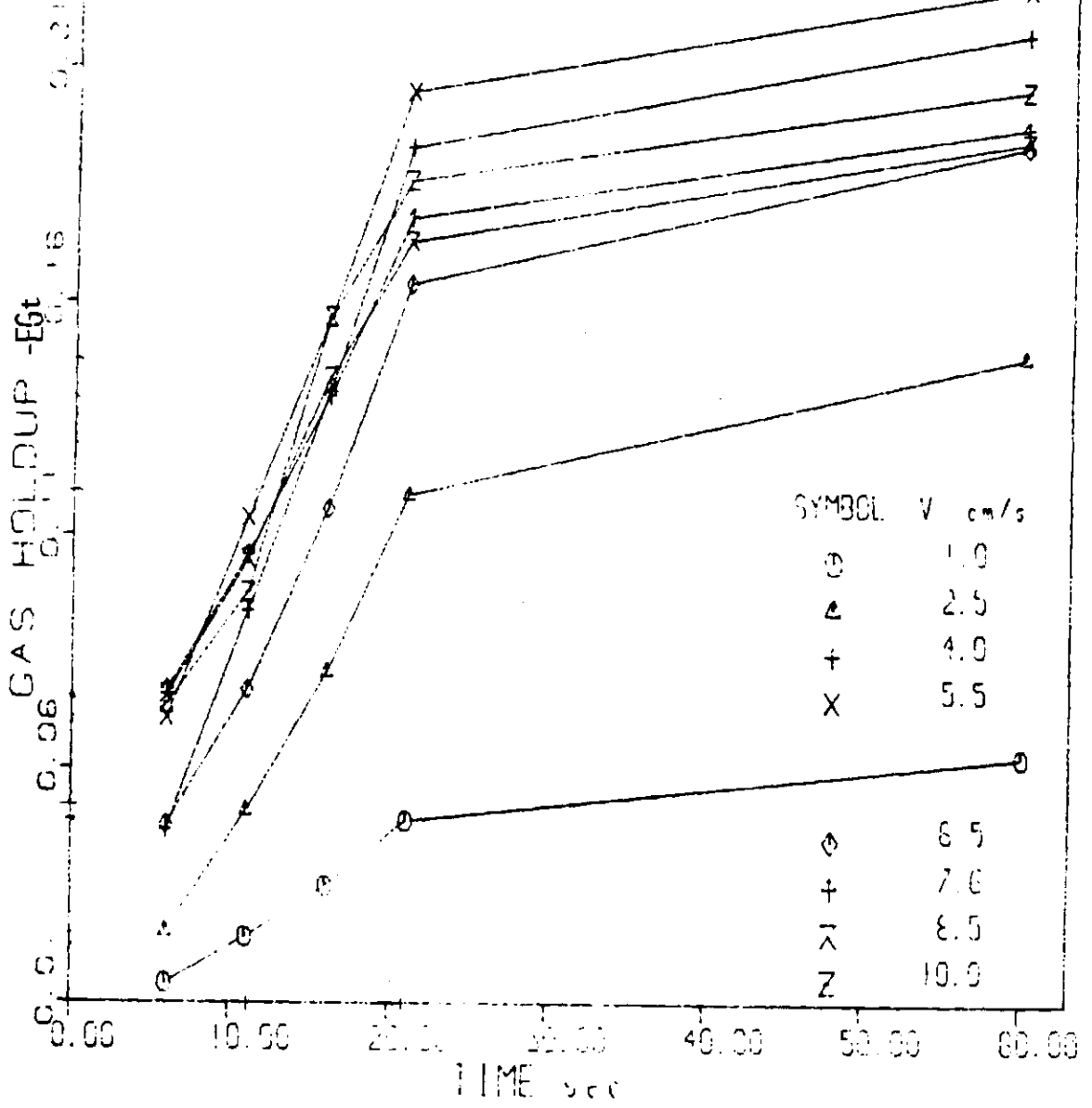
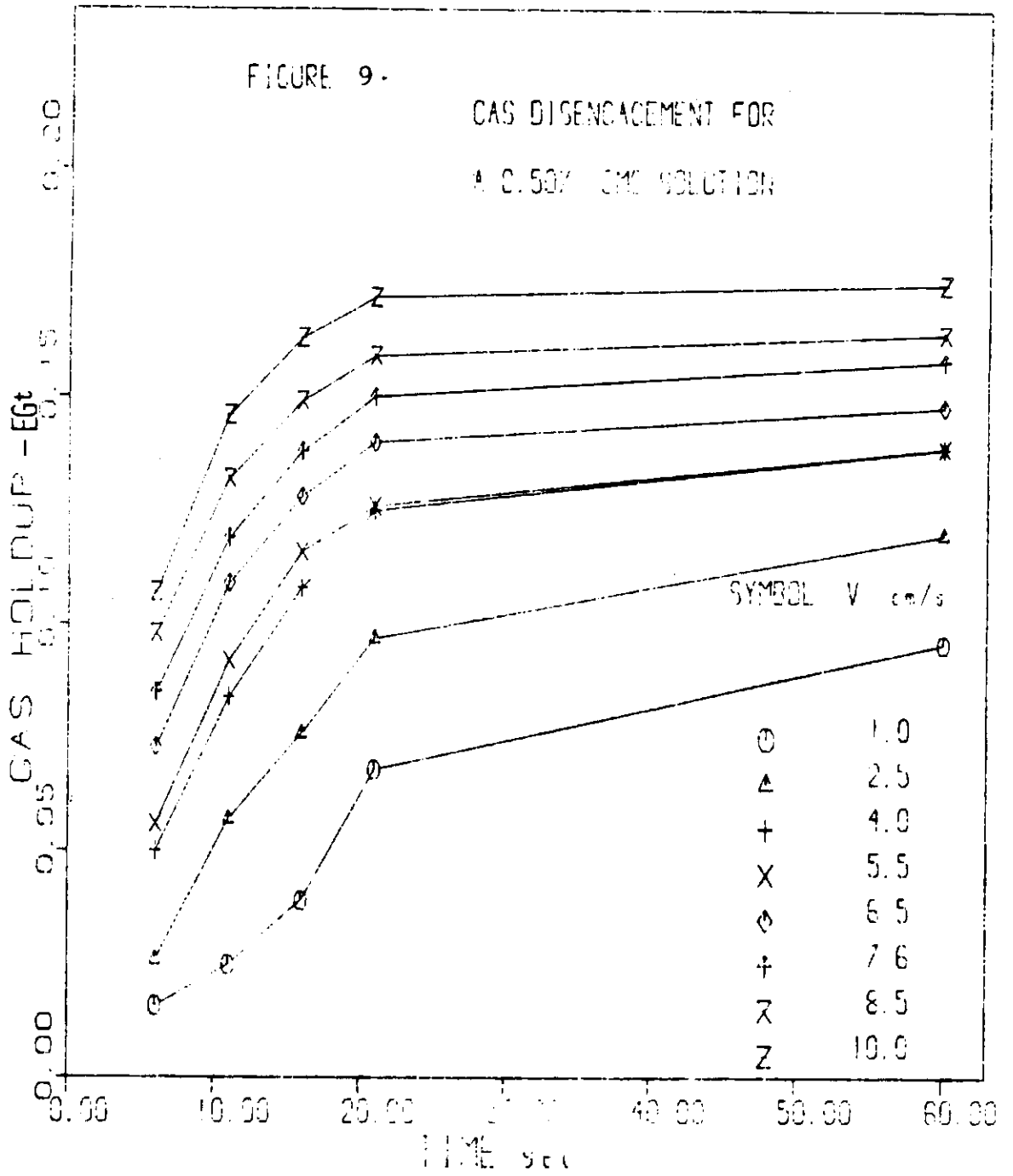
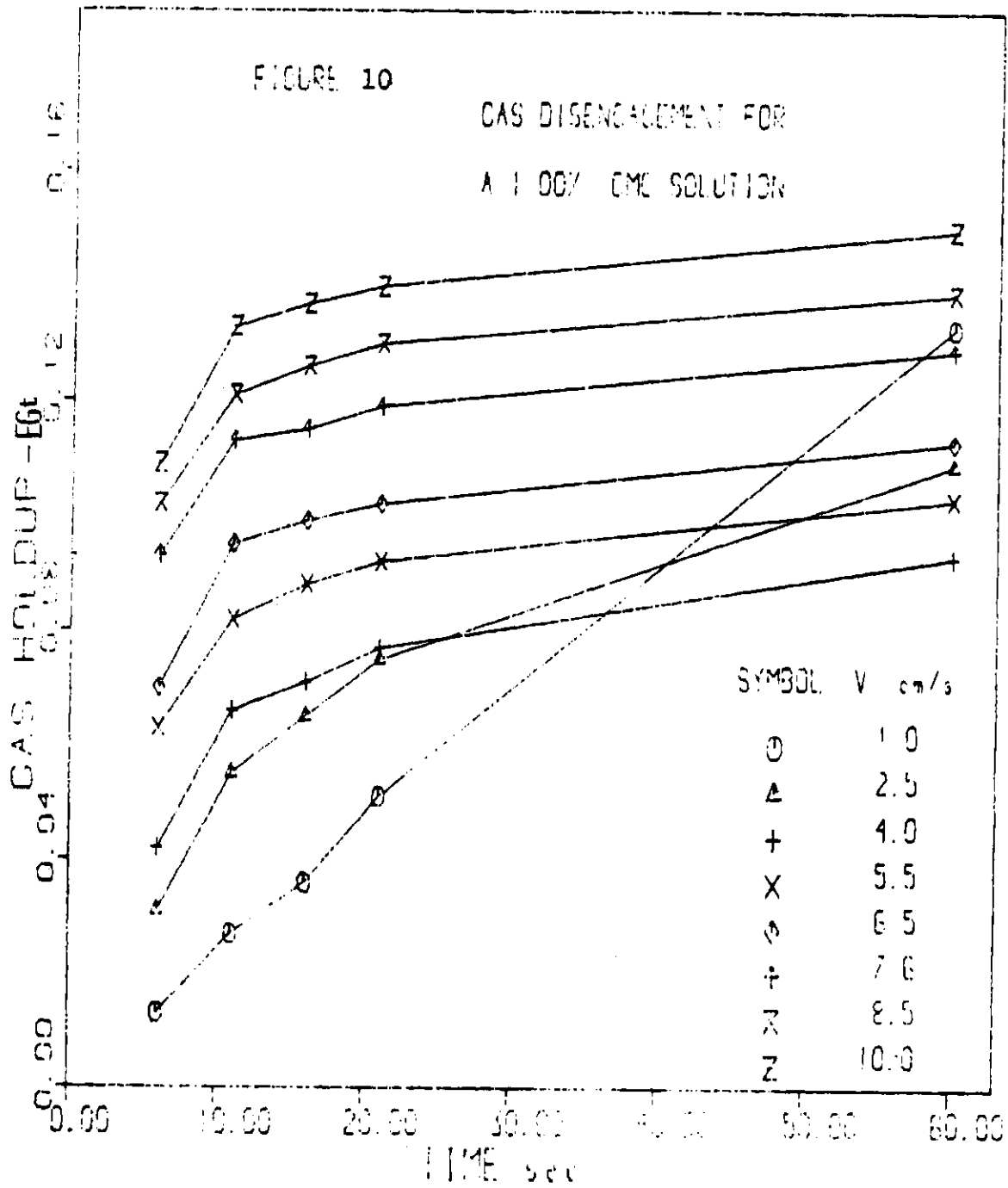


FIGURE 9.

GAS DISENGAGEMENT FOR  
A 0.50% CMC SOLUTION





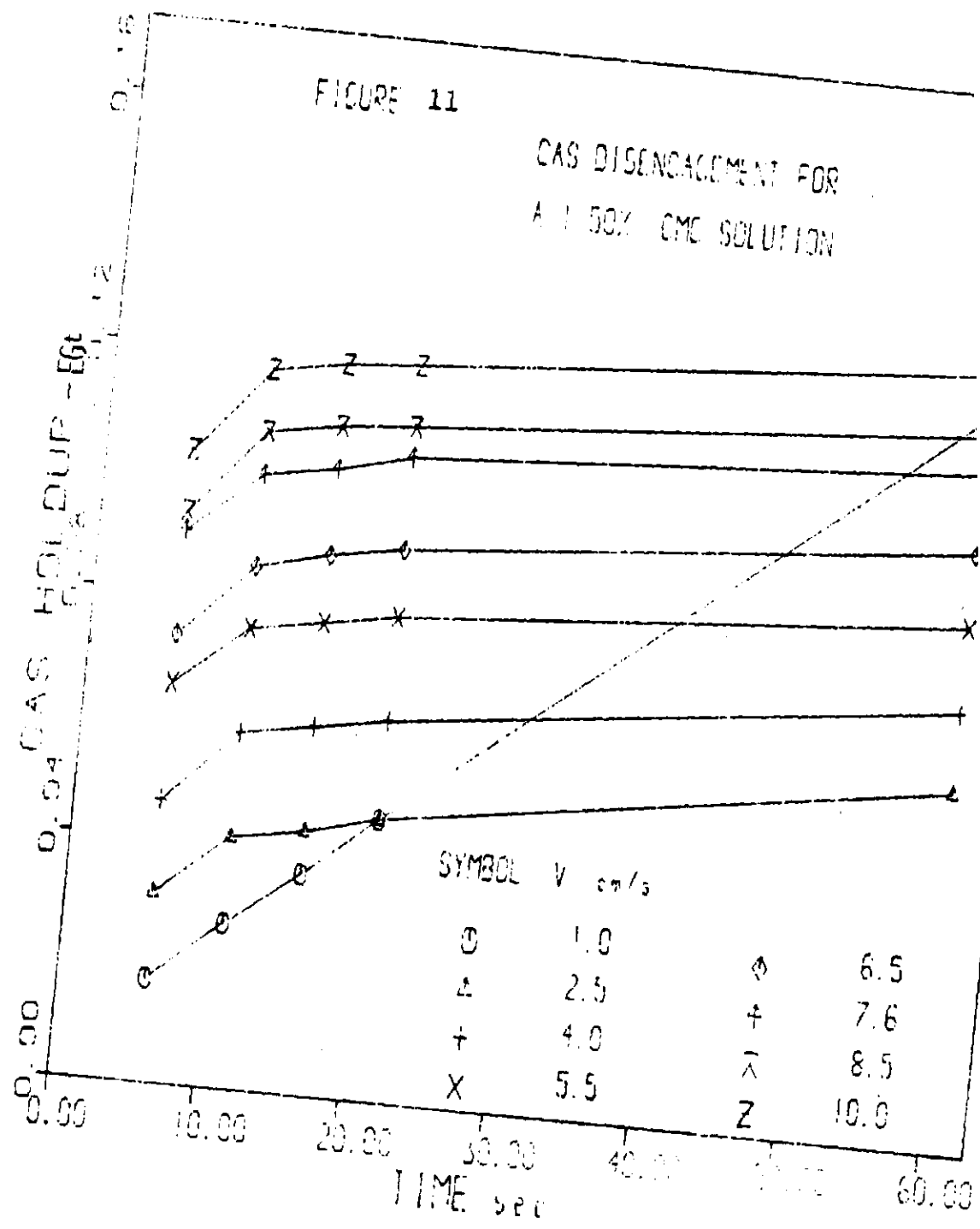
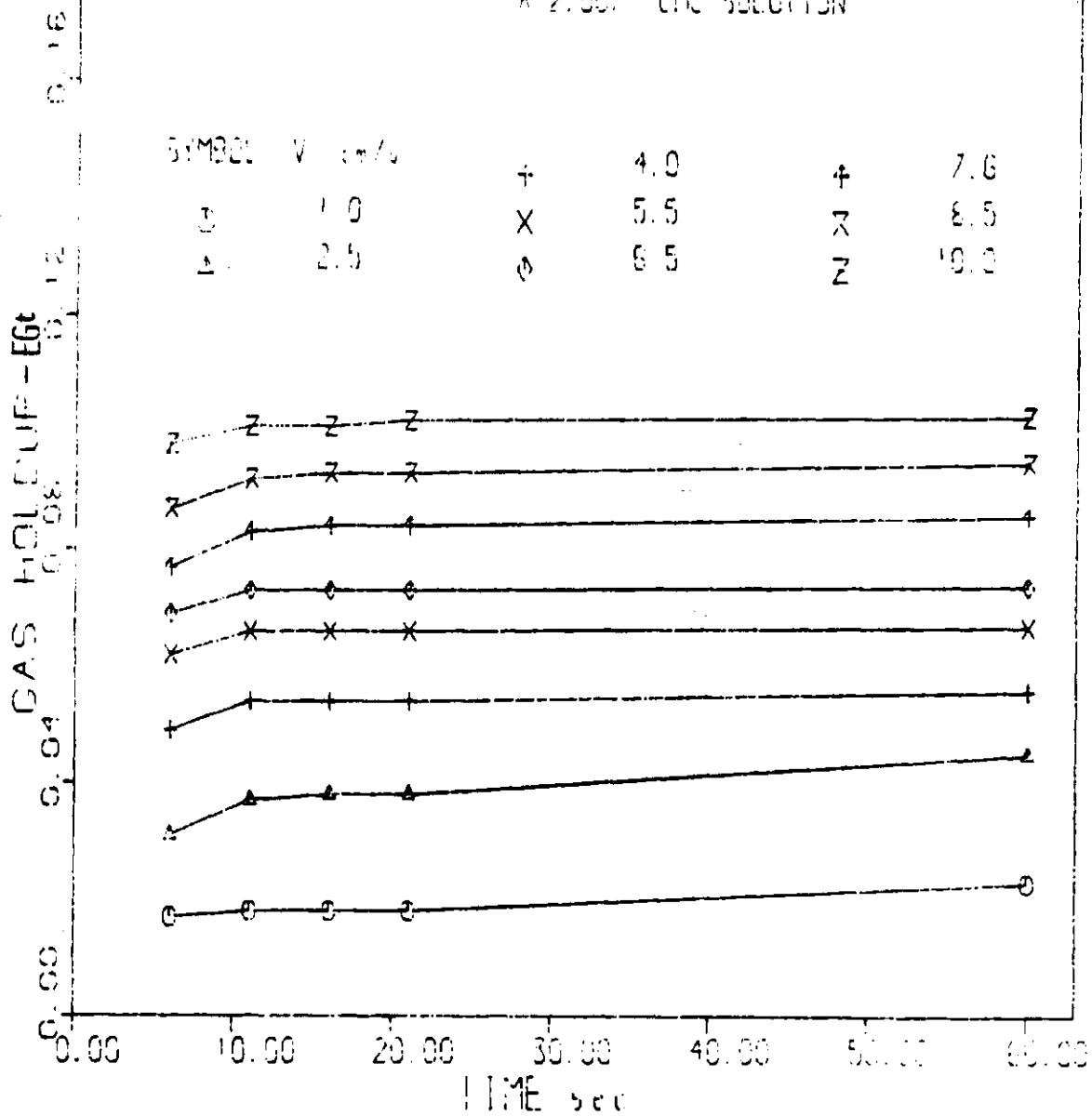


FIGURE 12

GAS DISENAGEMENT FOR  
A 2.00% CMC SOLUTION



# CALCULATED VS. OBSERVED HOLDUP

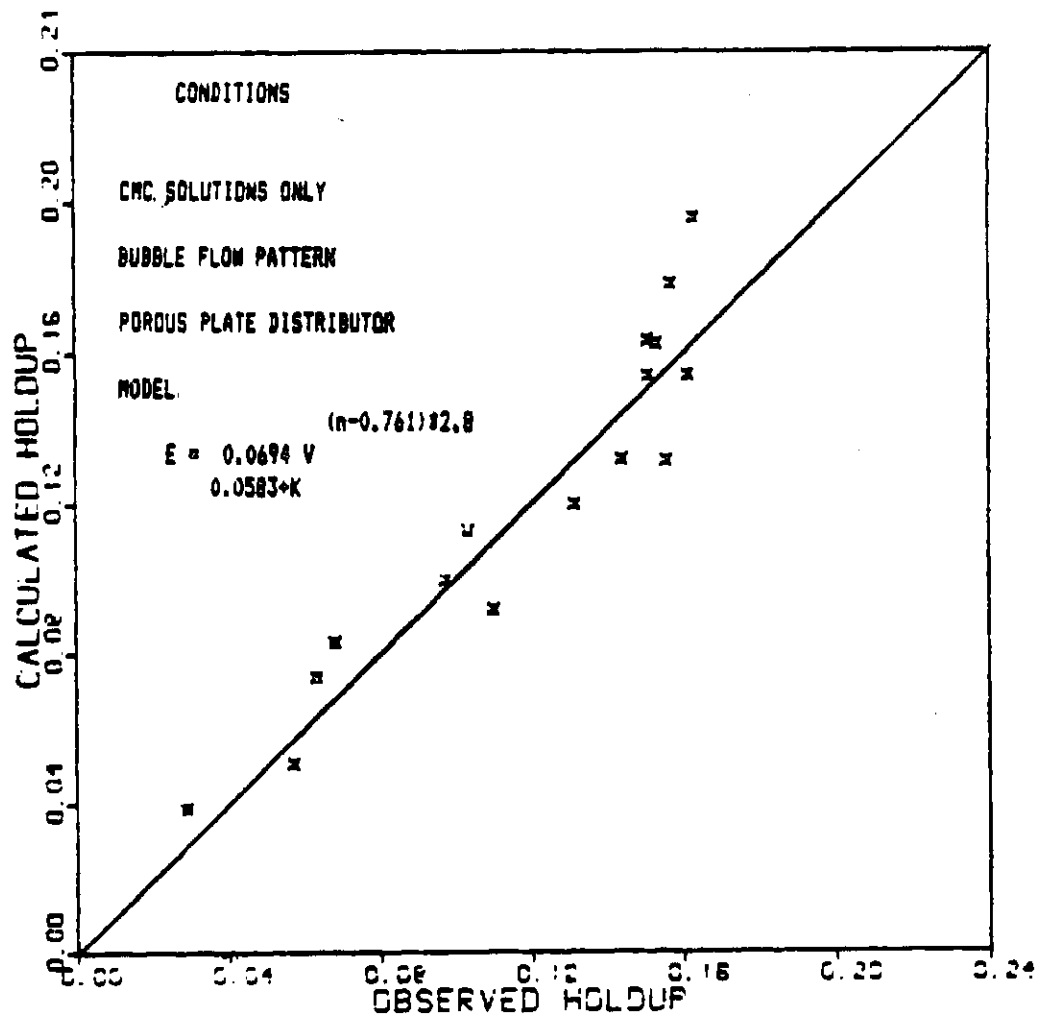


Figure 13 Parity Plot for Equation

APPENDIX- A

Holdup correlation for superficial gas velocity at gas inlet ( $E_G$  vs VGSO).

Mass Flow Rate  $W_G = VGS \times A \times \rho_{ATM} =$

Assume air as ideal gas and isothermal

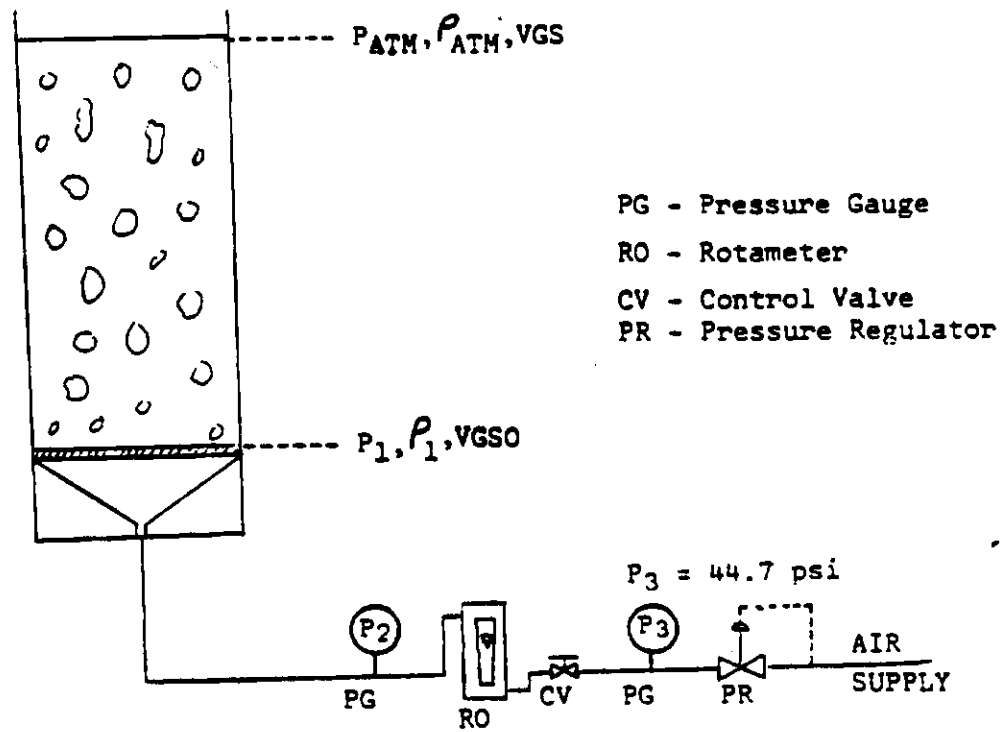
$$\frac{P_{ATM}}{\rho_{ATM}} = RT \quad \frac{P_1}{\rho_1} = RT$$

$$\frac{P_{ATM}}{\rho_{ATM}} = \frac{P_1}{\rho_1}$$

$P_1 = \text{hydrostatic head} = \rho_w g(1-E_G)H + P_{ATM}$

$$\rho_1 = \frac{\rho_{ATM}}{P_{ATM}} \times P_1$$

$$VGSO = \frac{W_G}{\rho_1 \times A}$$



LINE DIAGRAM

Sieve Plate

Correlation

$$E_G = 0.8034 (VGSO)^{0.5375}$$

Table for Sieve Plate-I

$P_2$ psi	$P_1$ psi	$W_G$ kg/s	$H_s$ m	$\rho_l$ kg/m <sup>3</sup>	VGSO m/s	$E_G$
24.70	21.49	0.036E-02	4.76	1.768	0.0111	0.068
24.70	21.31	0.063E-02	4.64	1.754	0.0196	0.092
24.70	21.15	0.082E-02	4.52	1.741	0.0261	0.115
24.70	21.06	0.101E-02	4.46	1.733	0.0320	0.126
24.70	20.98	0.120E-02	4.41	1.727	0.0380	0.137
24.70	20.87	0.148E-02	4.33	1.717	0.0474	0.153
24.70	20.76	0.177E-02	4.25	1.708	0.0568	0.169
25.20	20.54	0.237E-02	4.09	1.690	0.0769	0.199
25.20	20.34	0.294E-02	3.96	1.674	0.0962	0.225
26.70	20.01	0.391E-02	3.72	1.647	0.1302	0.271

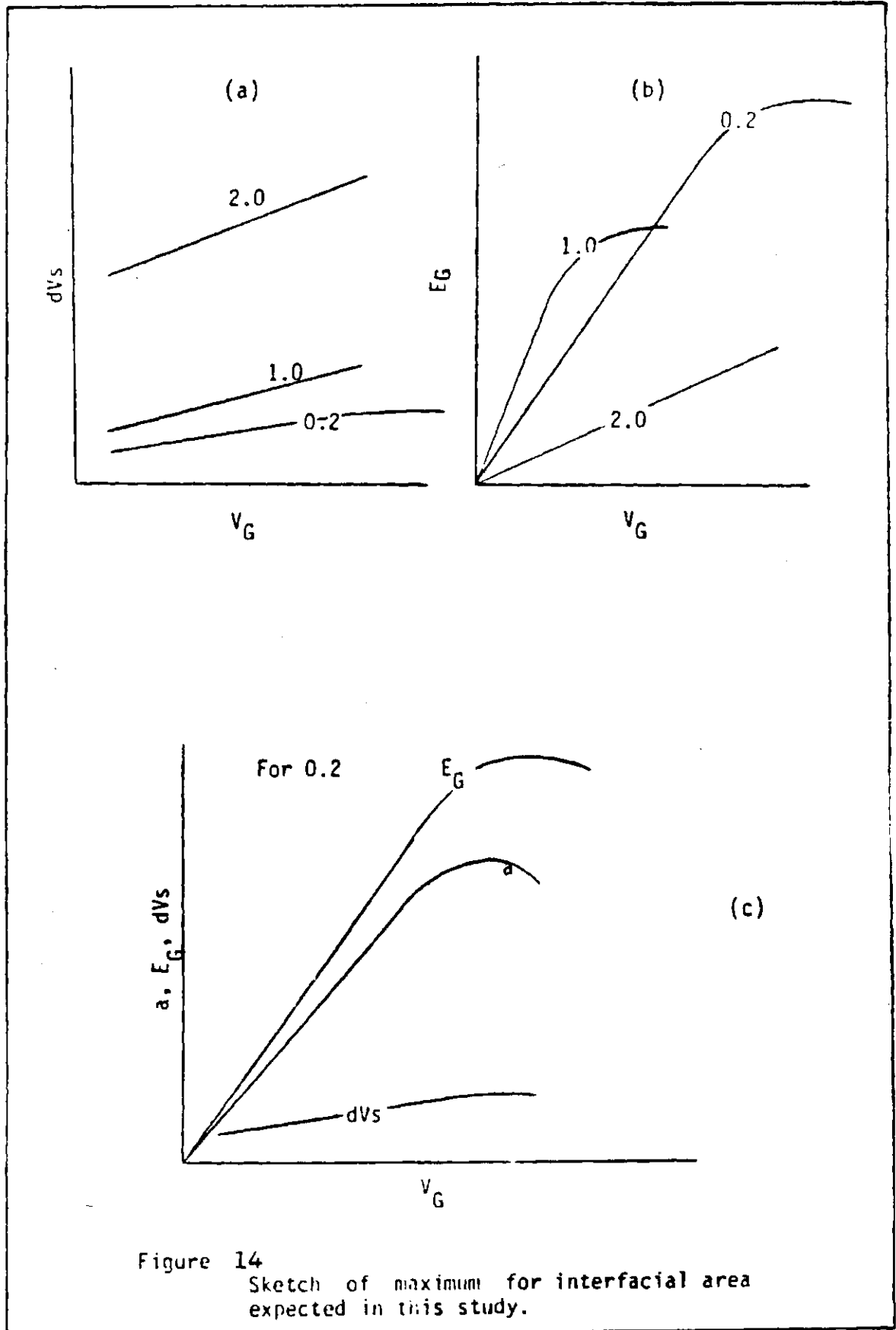


## APPENDIX A:

### Maximum in interfacial area

A major contribution of this study is the indication of a maximum in interfacial area in the bubble pattern. Interfacial area is important in determining rates of heat and mass transfer.

The justification of the presence of this maximum is depicted in the sketches of figure 14. Sketch 14(a) indicates linearly increasing bubble diameters with superficial gas velocity. Gas holdup increases and reaches a maximum as superficial gas velocity is increased. See figure 14(b). The exception to this, as shown, is at high concentration of CMC. The solution becomes viscous enough to promote coalescence and retard break-up thus reducing gas holdup and eliminating the maximum. If we use the standard relation, that interfacial area is directly proportional to gas holdup and inversely related to bubble diameter, we obtain a maximum in interfacial area. See sketch 14(c). Thus, operation of bubble columns, with porous plates, in the bubble pattern, at low enough viscosities, should result in maximum interfacial area.



## APPENDIX B:

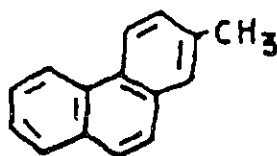
### Direct Coal Hydroliquefaction-Process Conditions and System Properties

A typical coal hydroliquefaction process (SRC-I,II) (4,81) is carried out at about 2000 psi (130atms.) and 450<sup>o</sup>c. Coal slurry, composed of fine coal powder (30 wt.%) suspended in coal derived liquids, and hydrogen is passed through a preheater, into the reactor-dissolver. Low residence time and high turbulence is maintained in the preheater to heat the three phase mixture at high heat transfer rates, to avoid coking. But the upper limit of the velocity of the three phase mixture is dictated by erosion considerations of the furnace tubes, due to the flow of coal slurry (41). The solvent and the heating are used to facilitate the thermal degradation of coal, resulting in the formation of free radicals of relatively low molecular weight. These free radicals are then stabilized by hydrogen transfer from the hydroaromatic solvent (41). High pressure of hydrogen and the inorganic matter "ash" present as residue, acts as a catalyst, to rehydrogenate the vehicle solvent. The role of such catalysts is important in the desulfurization, denitrogenation and other reactions leading to the formation of lighter products. This is in essence the SRC-II process.

The liquified product is generally aromatic in nature. Typical model compounds chosen as a representative of the product, used for determining the liquid phase physical properties, are illustrated on the next page.

Model Compounds

Wt.% of Compound



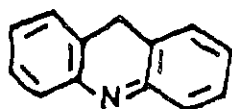
1-Methyl Phenanthrene

90-95%



Dibenzothiophene

3-5%



Acridine

2-5%

Physical properties of the coal slurry-hydrogen system will be evaluated on the basis of products and the earlier specified operating conditions. The various physical properties required, to draw the flow maps, are listed below.

1. Gas density
2. Liquid density
3. Liquid viscosity
4. Interfacial tension

1. The gas density is calculated by assuming that the gas is almost pure hydrogen with traces of hydrogen sulfide, methane and other organic vapors. A generalized compressibility chart corrected for hydrogen (110), is used to calculate the density at 2000 psia, 450<sup>o</sup>c. See Table I.

2. Liquid density: It is assumed the coal solvent slurry behaves like an ideal homogeneous mixture (pseudo homogeneous dispersion, discussed earlier). On this basis weighted average slurry density is calculated.

Density of coal-average of these reported values (105).

<u>Bulk Density of:</u>	<u>Lb/Cu-Ft</u>	<u>Gms/cc</u>
Anthracite	50-58	0.80-0.93
Bituminous	42-57	0.67-0.91
Lignite	40-54	0.64-0.86

Average density of coal-0.80 gms/cc

Density of solvent-average value of the reported densities (105) of the above described constituents of liquified coal. See Table III.

Average density-1.16 gms/cc

Concentration of coal in slurry, the commonly reported values lie between-25 to 35 wt.%.

Average concentration of coal in slurry- 30 wt.%

Average density of coal slurry- 1.067 gms/cc

3. The liquid viscosities cannot be evaluated accurately for lack of required data, thus the various values obtained from the literature are reported (92,132). See Table II.

4. Surface tension is calculated by assuming that the solid coal particles are small in size and light enough not to affect the surface tension. But the calculated surface tension is for room temperature and standard atmospheric pressure. Effect of temperature, pressure and hydrogen dissolution, cannot be accounted for, using the available simple theoretical equations (105). The calculated and reported values (69,132) of all of the four properties, are included in Tables I and II.

Before going directly to the study of the reactor hydrodynamics it would be worth looking into the reactor and preheater configurations, the reported sizes and estimated scale ups. Fig. 15 illustrates the reactor. The typical size of the preheater and reactor for the Wilsonville, SRC-I pilot plant are given below (85).

Preheater diameter: 1.5"

Preheater length: 130'

Transfer line diameter: 1.5"

Transfer line length: 123'

Reactor-dissolver diameter: 12"

Reactor-dissolver length: 23'

Stoichiometric requirement indicates that per unit volume of the slurry, about seven unit volumes of hydrogen would be required (69), at the desired operating conditions. This means that the superficial gas velocity should be somewhere between five to ten times that of the superficial velocity of the coal slurry, for the system to operate without appreciable recycling of either the gas or liquid phase. Table I reports the ranges of the velocities, experimented with, in SRC-II, EDS and H-Coal processes. The reported values of superficial velocities, for the SRC-I plant at Wilsonville are given below.

$U_{gs} = 1.8 \text{ cm/sec}$

U1s = 0.335 cm/sec

On the basis of these values and the expected turn over of the SRC-II process plant utilizing about 333,500 tons of coal/day (138), the estimated size of a single coal liquefaction unit, is something like 9 to 12 feet in diameter and 200 feet long. The L/D ratio is not more than 23 in this case, as discussed earlier, and hence the entrance region problems are anticipated.

With this brief description of coal liquefaction processes and the various parameters involved, we go over to evaluation of reported flow maps and generation of suitable flow maps, based on the available models.

By these techniques the properties used at 450°C and 2000 psi pressure for operating direct coal liquefaction reactor are

Density of Hydrogen = 0.0045 gms/c.c

Density of coal slurry = 1.15 gms/c.c

Viscosity of coal slurry = 1.0 - 100 cp

Surface tension = 20-30 dynes/cm

Table I

Physical properties of coal slurry hydrogen system  
 Operating conditions- 2000 psi(130 atms.), 450 c

Property	Calculated Values	Reported in Literature	Values used for this study
Density of hydrogen	0.287 lb/ft 0.0045 gms/cc	0.287-0.3 lb/ft	0.287 lb/ft 0.0045 gms/cc
Density of Coal Slurry	1.067 gms/cc	1.0-1.15 gms/cc	1.15 gms/cc
Viscosity of Coal Slurry	—	0.6-1.0 cp	1.0-100 cp
Surface Tension	54.28 dynes/cm at room Temp.	2.0-30 dynes/cm at Process Conditions	20-30 dynes/cm



Table II

Summary of reported operating conditions and properties

Process	Temp.	Press.	Slurry Density	Slurry Viscosity	Hydrogen Density	Superficial Velocities	
						Gas	Liquid
Solvent Refined Coal SHC-II	445 to 465 °c	1900 to 2000psi	1.15 gm/cc	—	0.287 lb/ft	0.045 ft./sec	0.0034 ft./sec
xxon Lonor Solvent EDS	370 to 480 °c	1450 to 2550psi	0.63 to 1.0 gm/cc	0.2 cp	0.3 lb/ft	0.2 to 0.5 ft./sec	0.02 to 0.04 ft./sec
H-Coal Amoco-HRI Process	427 to 482 °c	2000 to 3200psi	0.6 to 0.7 gm/cc	0.2 to 0.5 cp	—	0.04 to 0.16 ft./sec	0.07 to 0.18 ft./sec

Table III

Properties of representative components of coal derived liquid

Compound	Wt. %	Molecular wt.	mp °C	bp °C	Density gm/cc	Structural Parachor	Surface Tension dynes/cm
1-Methyl Phenanthrene	90-95	192.26	118	358	1.15	457.90	55.70
Dibenzothiophene	3-5	184.26	97	332	1.179	402.40	39.16
Acridine	2-5	179.22	110	346	1.16	402.90	43.98

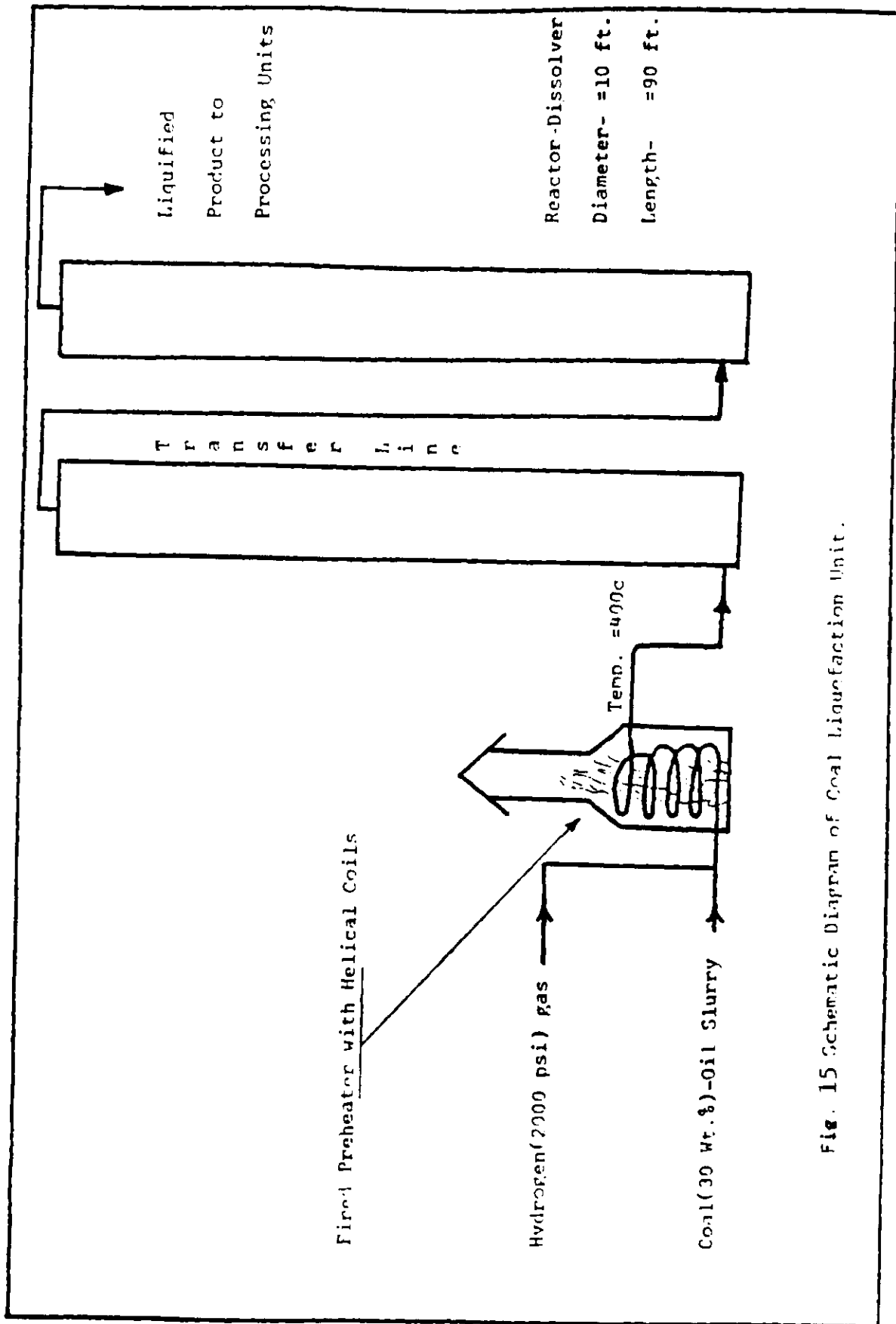


Fig. 15 Schematic Diagram of Coal Liquefaction Unit.

### APPENDIX C:

The derivation for interfacial area for bimodal distribution is given below:

Interfacial Area for bimodal distribution:

In any moment there are  $N_1$  small bubbles and  $N_2$  large bubbles present in the column for bimodal distribution. Large bubbles rise faster than small bubbles, then in any time interval  $t$ ,  $N_1$  small bubbles contacted the liquid in the column and  $N_2 V_2 / V_1$  large bubbles did too so, the total interfacial area is defined by:

$$A_T = A_1 + A_2$$

Where  $A_1$  and  $A_2$  are the contributions for interfacial from small and large bubbles respectively.

Substituting  $A_1$  and  $A_2$  by their geometric definitions (per unit of total volume):

$$= \frac{N_1 \bar{V} D_1^2}{V_T} + \frac{N_2 V_2 \hat{N} D_2^2}{V_1 V_T}$$

In the other hand, gas holdup for each of the sizes becomes:

$$EG_1 = \frac{\bar{N} D_1^3 N_1}{6 V_T}$$

$$EG_2 = \frac{\hat{N} D_2^3 N_2}{6 V_T}$$

Substituting for  $V_T$  in  $A_T$ :

$$A_T = \frac{6 EG_1}{D_1} + \frac{6 EG_2 V_2}{D_2 V_1}$$

Where  $V_2/V_1$  is found as the ratio of slopes for gas disengagement between 0 to 5 sec and 5 to 60 sec, representing large and small bubbles respectively (48).  $D_2$  is determined as a visual estimate of the average

equivalent diameter of large bubbles.  $D_1$  is determined from evaluating photos using the mean Sauter diameter for small bubbles. Gas holdup EG2 can be taken from Gas Disengagement figures at 5 sec and gas holdup EG1 the difference between total holdup and EG2.

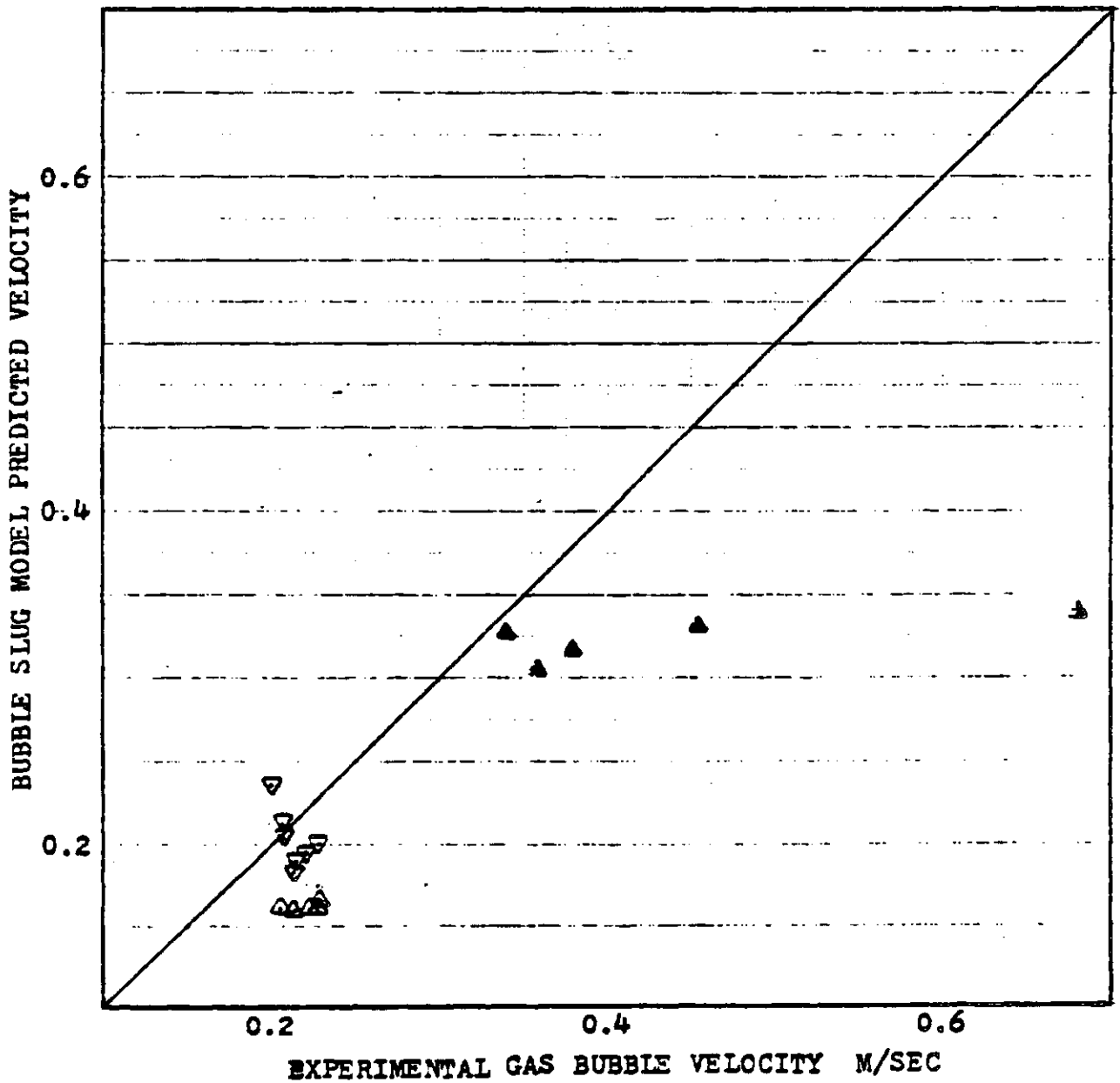
APPENDIX D

MODELLING CALCULATIONS

Figure 16.

COMPARISON OF GAS BUBBLE VELOCITIES

BUBBLE SLUG MODEL PREDICTED VERSUS EXPERIMENTAL



APPENDIX E

FLOW PATTERN TRANSITION CORRELATIONS

A. Bubble to Bubble-Slug:

$$U_{SL} = 3.0 U_{SG} - 1.15 \left[ \frac{\epsilon \sigma (\rho_L - \rho_G)}{\rho_L^2} \right]^{1/4}$$

$$U_{SL} = 2.2 U_{SG} - 1.128 \left[ \frac{\epsilon \sigma (\rho_L - \rho_G)}{\rho_L^2} \right]^{1/4}$$

B. Bubble to Finely Dispersed Bubble:

$$U_{SL} + U_{SG} = 4.0 \left[ \frac{D^{0.42} (\sigma / \rho_L)^{0.089}}{\nu_L^{0.072}} \left[ \frac{\epsilon (\rho_L - \rho_G)}{\rho_L} \right]^{0.446} \right]$$

C. Finely Dispersed to Slug or Churn:

$$U_{SL} = 0.923 U_{SG} - 0.734 \left[ \frac{\epsilon \sigma (\rho_L - \rho_G)}{\rho_L^2} \right]^{1/4}$$

D. Slug to Churn:

$$\frac{L_e}{D} = 40.6 \left[ \frac{U_{SL} + U_{SG}}{\sqrt{\epsilon D}} + 0.22 \right]$$

E. Slug or Churn to Annular:

$$U_{SG} = 3.1 \left[ \frac{\epsilon \sigma (\rho_L - \rho_G)}{\rho_G^2} \right]^{1/4}$$

F. Bubble-Slug to Slug or Churn:

$$U_{SL} = 1.79 U_{SG} - 0.286 \left[ \frac{\epsilon D (\rho_L - \rho_G)}{\rho_L} \right]^{1/4}$$

## APPENDIX F:

### PUBLICATIONS RESULTING FROM PROJECT

Kirpekar, A.C. Application of the Hydrodynamics of Gas Liquid Upflow Studies to Direct Coal Liquefaction Reactors. Kingston, RI: University of Rhode Island; 1980. Dissertation.

Kamat, A.B. Gas Holdup and Flow Pattern Studies in a Bubble Column. Kingston, RI: University of Rhode Island; 1981. Dissertation.

Mikkilineni, S. The Effect of Gas Distributor on Holdup and Flow Pattern in a Bubble Column. Kingston, RI: University of Rhode Island; 1983. Dissertation.

Trigueros, C.E. Interfacial Area Analysis of a Non-Newtonian Liquid and Air Flowing Cocurrently up in a Bubble Column. Kingston, RI: University of Rhode Island; 1983. Dissertation.

Sullivan, M.D. Holdup and Flow Patterns of Non-Newtonian Fluids in Bubble Columns. Kingston, RI: University of Rhode Island; 1983. Dissertation.

Turgeon, E.J. A Hydrodynamic Model for Co-current Gas-Liquid Bubble-Slug Flow in Vertical Tubes. Kingston, RI: University of Rhode Island; 1984. Dissertation.

Bouressa, J., Experimental Studies and Modelling of the Slug Pattern in Two Phase Non Newtonian Flow. Kingston, RI: University of Rhode Island; Dissertation, Expected 1984.

Holcombe, N.T.; Smith, D.N.; Saroff, L.; O'Dowd, W.; Knickle, H. Thermal Dispersion and Heat Transfer in Non-Isothermal Bubble Columns. AIChE Annual Meeting, New Orleans; 1981 November.

Knickle, H.N.; Kirpekar, A.; A Proposed Flow Map in the Design of Coal Liquefaction Bubble Column Reactors. AIChE Winter Meeting, Orlando; 1982 February.

Knickle, H.N.; Kamat, A. Prediction of Hydrodynamic Effects in Bubble Column Type Coal Liquefaction Reactors. AIChE Annual Meeting, Los Angeles; 1982 November.

Knickle, H.N.; O'Dowd, W.; Holcombe, N.T.; Smith, D.N.; Backmixing and heat Transfer Coefficients in Bubble Columns Using Aqueous Glycerol Solutions. 21st ASME/AIChE National Heat Transfer Conference; Seattle; 1983 January. (Also ChE Symposium Series)



## NOMENCLATURE

A	Cross sectional area of the pipe
a	Interfacial area
B,n	Constants
D	Column diameter
db	diameter of the bubble
de	Equivalent diameter
dh	Orifice or hole diameter
DVs	Volume-surface mean diameter
e	Energy dissipation
Eg,Ega	Average (overall) gas holdup
Eg(z)	Local gas holdup
g	Acceleration due to gravity
H	Height of column or bubbled bed
Hs	Liquid height without gas flow
KL	Liquid mass transfer coefficient
L	Length of the pipe
Le	Entrance length for the formation of slug flow
m	Reaction order
n	Power law index
NBo	Bond number ( $g D^2 \rho_L / \sigma$ )
NFr	Froude number ( $VG / \sqrt{g D}$ )
NGa	Galilei number ( $g D^3 / \nu L^2$ )
NWe	Weber number ( $\rho D VG^2 / \sigma g_c$ )
NEo	Eotvos number ( $g D e^2 \rho / \sigma$ )
P	Pressure
Q	Volumetric flow rate at operating conditions

$Ra^2$	Miller's adjusted multiple correlation coefficient
$r^*$	Dimensionless radius
$R_c$	Radius of curvature
$Re_b$	Reynolds for the bubble
$Re_h$	Reynolds in the hole or sparger
$U_o$	Velocity of a single large gas bubble
$V_{bf}$	Average bubble rise velocity
$V_{GS}$	Superficial gas phase velocity
$V_{GS0}$	Superficial gas velocity at gas inlet
$V_h$	Velocity in the hole
$V_{LS}$	Superficial liquid phase velocity
$V_s$	Slip velocity
$W$	Mass flow rate
$w_o$	Maximum vorticity
$Z$	Height (position)

### Greek Symbols

$\rho$	density
$\sigma$	surface tension
$\mu$	viscosity
$h_m$	difference in manometer reading
$h_w$	pressure drop in units of water
$z$	difference in height
$\nu$	Kinematic viscosity
$\mu_{app}$	Apparent viscosity
$\dot{\gamma}$	Shear rate
$\delta$	Porous size
$c$	Apparent viscosity at $dV/dy=0$

:

a	Apparent viscosity at $dV/dy=$
$\lambda$	Time constant in rheological characterization
$\Delta\rho$	Liquid-Gas densities difference
$S_i$	Partial cross sectional area
$\alpha$	Parameter in Kato's correlation ( $=1-e^{-0.2 V_G^2}$ )

Subscripts

a	air
b	bubble
e	entrance
G	gas
L	liquid
m	manometer fluid
M	mixture
o	orifice
w	water